Desulfurization of Low Temperature Char
II - Rate Data in A Continuous System

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INTRODUCTION

The desulfurization of low temperature char in batch fluidized systems was presented in previous publications[1]. Commercial application of the process would most likely require the use of a continuous processing technique. The assessment of the commercial potentialities of the process therefore requires that experimental data be obtained in a system which is operated continuously both with respect to the solids and the gas.

Such data have now been obtained in a continuous bench scale fluidized unit both using fresh hydrogen and recycle gas for desulfurization and are presented here. Inhibition by the hydrogen sulfide product was found to control the amount of desulfurization achieved. The results in this sense are similar to those reported previously in deep batch beds[2]. The data are interpreted in terms of the total inhibition isotherms discussed previously.

Subsidiary data of interest are also presented relative to the rate of production of methane by direct hydrogenation of char and to the yield of hydrogen produced by the devolatilization of char.

Prior data are available on the production of methane by hydrogenation of char by Dent[3] and others[4,5]. None of the published data were obtained, however, in a continuous system.

EXPERIMENTAL

Raw Material

A low temperature char produced by the fluidized carbonization of a Pittsburgh Seam coal at 950°F was used in this work. It was the same Arkwright char[2] previously used in our batch work. An average analysis of the feed char is given in Table VI. The char was fed without prior drying. The moisture and elemental analysis was determined for the feed char to each run. These analyses varied somewhat from run to run and each individual analysis was used in determining the elementary balances for the corresponding run. Only an average analysis with the exception of the sulfur content is given for the sake of brevity. One run was made, No. 58, with a char which had previously been desulfurized to 1.47 percent sulfur.

Equipment

The runs were all made in a continuous bench scale fluidized reactor system. The equipment had been designed to study gasification reactions at

temperatures up to 1800°F and pressures up to 50 atmospheres. The details of the equipment are quite complex and for brevity will largely be eliminated from the discussion here. A simplified flow diagram is shown in Figure 1.

The char was fed from a pressurized hopper by a rubber roll feeder. The feed rate is regulated by a variable speed drive which turns the feeder shaft. A pressure equalizing line equalizes the pressure between the feed hopper and the discharge from the rolls.

The solids are picked up with the process gas, i.e., either fresh hydrogen or recycle gas and transported into the fluidized reactor.

The hydrogen was passed through a purifying train for removal of oxygen and moisture which consisted of a nickel-alumina catalyst chamber and a silica gel tower in series. These are not shown. The hydrogen was then metered into the reaction system by means of a rotameter. The recycle gas runs were conducted by recycling the product gas, after removal of hydrogen sulfide, with a Gast rotary pump enclosed within a pressure housing. The hydrogen sulfide was removed by passing through a catalyst tube packed with 1/8-inch pellets of a sulfur acceptor containing 80% Cu-10% V-10% Cr. The preparation and regeneration of this material was carried out according to the procedure recommended by the U. S. Bureau of Mines[8].

The fluidizing vessel was a 4-inch diameter cast Duralloy (28% Cr-4% Ni) tube approximately 55 inches long with a 60 degree cone bottom. It was supported by means of a ring joint flange to an upper section which was attached to the top flange. This particular alloy was found to be quite resistant to attack by hydrogen sulfide.

The reactor tube was surrounded by a concentric furnace which contained four independently controlled heater circuits. The reactor tube was maintained in pressure balance with the furnace jacket which were both enclosed in a large water jacketed pressure vessel.

The pressure bearing walls were held at 450°F by controlling the steam pressure and water level in the jacket.

The inlet line to the reactor tube passes through a packing gland at the bottom of the shell vessel and thus seals the furnace zone from the reaction zone. Pressure balance was maintained across the reactor by using a Kendall gas relay to feed nitrogen to the furnace zone. Purge nitrogen was also fed at a metered rate through the top reactor flange to purge the dead space above the bed level. This was replaced with recycle gas in the recycle gas runs.

The fluidized bed level was controlled by an adjustable J-tube which entered through a packing gland in the top flange. The solid and gaseous reaction products left the reactor through the J-tube and into a 1-1/2" internal diameter cyclone. The solids knocked out of the gas in this cyclone flow by gravity to either of two char receivers selected by the char diversion valve mounted above them. The partially cleaned gas passed on to either of two 1/2" internal diameter cyclones each of which were mounted inside a fines receiver. A fines diversion valve was placed upstream for switching from one receiver to another. Final cleanup of the exit gas from the fines receiver was accomplished by the dust filter which was packed with a Fiberglas blanket material.

The filtered gas is cooled and the moisture condensed. A liquid letdown valve allows removal of the condensate.

Excess gas was throttled through a back pressure control valve and through a water bubbler to the wet test meter. This valve automatically controlled the system pressure. Part of the exhaust gas stream could be by-passed around the bubbler and through the thermal conductivity cell.

The fresh hydrogen runs were conducted on a once-through basis with respect to gas. After temperature and pressure in the reactor had lined out, the char feed was started and continued long enough to fill the reactor and pass three additional inventories through the reactor. The material balance was initiated by switching the char and fines solid products from the lineout pots to the material balance pots. After the material balance period, the unit was switched back to the lineout catch pots and shut down immediately by cutting of both the feed char and gas.

The bed of char remaining was drained from the reactor and determined as the inventory weight. The weight of char fed was determined by difference between the amount charged and the amount left in the feed hopper after the run. The residence time was calculated by comparison of the product rate to char inventory weight.

Gas samples were taken during the material balance period of the run. Triplicate bottles were used for a gravimetric[7] analysis for carbon dioxide and total carbon and hydrogen content, infrared analysis for methane, ethane and carbon monoxide and a Tutweiler analysis for hydrogen sulfide.

Moisture yields were determined by the difference between the condensate yield and the moisture in the feed char after allowing for uncondensed moisture in the gas.

Ultimate analyses were obtained on the product and feed chars.

The recycle gas runs were conducted in a similar fashion. The unit was started in operation with the system full of hydrogen. The lineout period in these runs was continued until the outlet gas composition reached a steady value as indicated by the thermal conductivity cell. Adjustment of the recycle gas rotameter reading was effected during the lineout period to compensate for changes in gas density.

The operating data for the fresh hydrogen and recycle gas runs are given in Tables I and II. The material and elementary balances and yields for the fresh hydrogen runs are given in Table III. A negative figure indicates a deficiency in the product as compared with the feed. The yield figures are given as determined with the exception of the char yield which is forced to give a 100 percent weight balance. Char yield is omitted where the discrepancy in the weight balance was greater than 3 percent. The gas and solids analyses are omitted for the sake of brevity.

Table IV gives the corresponding data for the recycle gas runs. All the yields in this case are forced, i.e., the char yield is adjusted to give a 100 percent closure of the weight balance while simultaneously the liquor and product gas rates were adjusted to obtain closures around the oxygen and hydrogen balances, respectively.

Tables V and VI give the product gas and char analyses for the recycle gas runs. The gas analysis is recorded as determined with the exception of the nitrogen content which was forced to give a nitrogen balance.

INTERPRETATION OF RESULTS

Discussion in Terms of Total Inhibition Isotherms

The conditions which correspond to total inhibition of the desulfurization process have been given in the previous paper[2] in the form of the corresponding isotherms. These were used to determine the hydrogen circulation requirements in batch fluidized systems. The isotherms may also be used directly to determine the "equilibrium" sulfur reduction that would be obtained in a single stage continuous fluidized system.

The hydrogen circulation requirements calculated in this manner are shown as a function of the sulfur level of the product char in Figure 2. Curves are shown for three different treatment temperatures, i.e., 1100°F, 1350°F and 1600°F. The hydrogen requirements are given in terms of standard cubic feet per pound of dry feed char measured at 70°F. The curves are based on the use of an Arkwright char feed which has an initial sulfur content of 2.45 weight percent.

The hydrogen requirements are considerably larger than those given previously for the batch system. This is because of the absence of the staging effect characteristic of the batch system. Staging in a continuous system may be effected by separately devolatilizing the char and then using a multi-stage countercurrent hydrogen treatment system for desulfurization.

The devolatilization reduces the sulfur of the feed char from 2.4 to 1.9 weight percent without the addition of external hydrogen. Calculated curves giving the hydrogen recirculation requirements for multi-stage systems of the above type at 1600°F are also shown in Figure 2. The curve for the ideal countercurrent system corresponds to the lowest hydrogen requirements even where a very large number of stages are used.

The method of calculation was a graphical one and similar to that used in conventional countercurrent adsorption calculations.

All of our experimental data were obtained in a single stage fluidized system. Some of our experimental points are shown for comparison with the calculated curves in Figure 2. Also a comparison is made in case of all the runs of the observed hydrogen sulfide to hydrogen ratio in the product gas with the ratio that corresponds to an "equilibrium" condition. These data are given in columns 10 and 11 of Tables I and II.

It is noted that only under the very mildest conditions used was the amount of desulfurization less than corresponds to "equilibrium", i.e., at hydrogen partial pressures of one atmosphere or less or at temperatures below 1350°F and residence times of the order of 20 minutes or less.

It is also seen that under the more severe conditions, i.e., at hydrogen partial pressures above about 3 atmospheres that the amount of desulfurization achieved was greater than corresponds to "equilibrium". Thus at the most severe conditions employed, i.e., total pressure 11 atmospheres, 28 minutes residence time and 1600°F the amount of desulfurization achieved was better than could be obtained under "equilibrium" conditions even if the char was devolatilized in a separate stage before desulfurization.

It is apparent from this violation of the "equilibrium" desulfurization concept that the sulfur in the raw char is in a more labile condition both in

the thermodynamic and kinetic sense than after the char had been thermally treated for several hours. The experimental data upon which the total inhibition isotherms are based unavoidably involved thermal treatment of the char for several hours. This undoubtedly causes a transformation of the sulfur to a more stable form. Such a phenomenon was previously noted by Powell[8] who ascribed it to a transition from the absorbed to solid solution states.

It is clear, from the above, that the hydrogen circulation requirements as given in Figure 2 are conservative and that better results can be obtained in practice at least under relatively severe operating conditions.

It is probable, however, that the equilibrium curves of Figure 2 would adequately predict the hydrogen requirements for the production of lower sulfur chars of the order of 0.6 weight percent or better. This is indicated by the results of Run No. 58. Here a char that had been previously desulfurized to 1.47 weight percent sulfur was further desulfurized at 1600°F and 6 atmospheres to produce a product char of 0.62 weight percent sulfur. The amount of desulfurization achieved corresponded quite closely to the "equilibrium" condition. These same experimental conditions were sufficiently severe, however, to produce a greater than "equilibrium" amount of desulfurization with the normal high sulfur feed char.

Kinetics of Desulfurization Process

The above data suggests that the net rate of desulfurization is determined by the competition between two processes, namely, the thermal fixation of the sulfur to produce a more stable form and its rate or removal by hydrogen while still in the labile form.

Since thermal fixation occurs more rapidly the higher the temperature it would not be surprising to find an optimum temperature for maximum desulfurization. Some of the differential rate data presented in the previous paper[2], i.e., Figures 6 and 7 suggests the possibility of an optimum rate at about 1450°F for desulfurization at low pressures of the order of one atmosphere.

The present data are not very suitable for kinetic interpretation since "equilibrium" conditions or better were achieved in most of the runs. An empirical approach to the question of the relative desulfurization rates at 1350 and 1600°F was therefore used.

The desulfurization efficiency is shown in Figure 3 as a function of an empirical severity factor. The desulfurization efficiency is defined as the ratio of the hydrogen sulfide concentration observed in the exit gas to that which would be obtained if "equilibrium" were achieved. The severity factor is rather arbitrarily defined as the partial pressure of hydrogen in the exit gas multiplied by the square root of the solids residence time.

These data illustrate the point discussed earlier that equilibrium conditions or better were achieved in most of our runs.

The bulk of the data were at the two temperatures of 1350°F and 1600°F, respectively. A statistical analysis of the data at these two temperature levels was made to discern whether any difference in desulfurization ease could be noted. A best straight line was drawn through the origin in

both cases by the method of least squares. The lines were forced through the origin for physical reasons since it is obvious that the desulfurization efficiency should be zero when the severity factor is zero.

A difference in slope between the 1350°F and 1600°F data is apparent. The analysis of variance was applied to these data to determine whether this difference in slope is a real effect or simply a result of the random scattering of the data. The t-test outlined by Youden[9] was applied and a value for t equal to 2.1 was obtained. Since the system has 16 degrees of freedom, a probability equal to 0.95 is found that the difference in slopes is real.

It is therefore concluded that less severe operating conditions are required to achieve equilibrium or better desulfurization at 1350°F than at 1600°F.

This is in accord with the above discussion on relative rates of thermal fixation and removal of sulfur.

Methane Yields and Approach to Equilibrium

The data are potentially of commercial interest to the problem of producing pipeline gas by partial hydrogenation of char at low pressures. It is of interest first of all to examine the data to determine how much methane is produced relative to the graphite hydrogen equilibrium.

The equilibrium constants $K = P_{H_2}^2/P_{CH_4}$ in the effluent gas from the 1350°F fresh hydrogen runs all, with the exception of Run 57, fall within the range of 19 to 53. Since the equilibrium constant is 10 at this temperature, it is clear that the methane content of all the fresh hydrogen runs at 1350°F fall short of the graphite equilibrium value.

Figure 4 shows the approach to the graphite equilibrium in the fresh hydrogen runs at 1600°F. In this case it is seen that the methane content of the gas is greater than the equilibrium value at pressures below 4 atmospheres. Even at the highest pressure studied, i.e., at eleven atmospheres the methane concentration approaches within 65 percent of the equilibrium value.

The high methane yield in relationship to the graphite equilibrium can be ascribed to the selective hydrogenation of the volatile matter of the char as was proposed previously by Dent[3]. Such a process would not be controlled by the graphite equilibrium.

The approach to graphite equilibrium in the recycle gas runs is illustrated in Figure 5. It is seen that the methane concentration at all temperatures studies is well in excess of the graphite equilibrium value. The methane in these runs is produced by thermal decomposition of the char volatile matter. The methane yield in such a process would also obviously not be controlled by the graphite equilibrium.

An examination of the data for the fresh hydrogen runs shows that the methane yield, everything else being equal, increases with the operating pressure. A somewhat higher yield on the average is obtained at 1350°F than at 1600°F.

The yield of methane in the fresh hydrogen runs is always greater than in the recycle runs even at low pressures. At 11 atmospheres the yield of methane is approximately 3.5 times greater in the fresh hydrogen runs.

Reaction Rate Constants

An earlier study[5] gave quite extensive information on the rate of hydrogenation of devolatilized char. It would be of interest to compare the present data on raw char with the previous data.

The data at 1600°F could be more easily compared since there are earlier data at the same temperature. Unfortunately, the close approach to methane equilibrium makes a kinetic analysis of these data relatively meaningless.

This problem does not exist with the 1350°F data and the rate constants are shown as a function of the amount of methane produced in Figure 6.

The rate constants are calculated on the assumption that the reaction rate is first order with respect to hydrogen pressure. It is also assumed that the gas in the fluid bed is perfectly mixed, i.e., the outlet partial pressure of hydrogen controls the reaction rate. Also in calculating the rate the yield of methane produced by devolatilization was subtracted from the observed methane yield.

For comparison purposes the rate constants for devolatilized char at the lowest temperature investigated, i.e., 1500°F, is shown.

It is seen that it is not possible to correlate all the experimental points over the whole pressure range. All the points at 3.0 atmospheres pressure lie approximately on a straight line. The line illustrates the rapid decrease in rate to a value more in line with that of devolatilized char as the amount of char hydrogenated to methane increases.

The higher pressure points are insufficient in number to point up any definite trends. However, the first order rate constants, at the same level of methane yield definitely increase with pressure. It may be conjectured, therefore, that the period of abnormally high hydrogenation rate extends to higher methane yields as the pressure is increased. This would have to be confirmed by more experimental data, however.

Gas Yields in Char Devolatilization

The devolatilization of char could conceivably be of commercial interest from the point of view of hydrogen production. It is noted in Table V, for example, that the gas produced by devolatilization at 1600°F contains better than 70 percent hydrogen or better than 80 percent of hydrogen plus carbon monoxide.

The yields of these gases as well as the hydrogen concentration increases rapidly with the temperature. The yield-temperature relationship is shown in Figure 7.

The relative constancy of the liquor yield and the increased yield of carbon monoxide with increasing temperature can be assumed to show increasing direct gasification with temperature of the char by the water vapor present in the system.

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Table I

Fresh Hydrogen Runs

Operating Data

Run No.	Dry Char Feed Rate Lbs./Hr.	Temp.,°F	Press. Atm.	P_{H_2} Atm.	Inlet H$_2$/Char SCF/Lb.	Res. Time Min.	% Sulfur in Char Feed	% Sulfur in Char Product	P_{H_2S}/P_{H_2} x 100) Obs. in Exit Gas	Equil.
54	7.26	1202	3.01	2.41	12.7	11	2.47	1.72	0.50	1.55
55	7.29	1202	3.01	2.34	12.6	23	2.43	1.69	0.89	1.34
51	6.91	1206	3.01	2.40	13.6	36	2.33	1.61	0.82	0.86
14	3.87	1350	1.28	1.08	15.2	21	2.46	1.60	0.87	1.24
16	3.73	1361	1.21	1.04	15.6	78	2.45	1.43	0.71	0.66
23	6.01	1348	1.40	0.99	5.9	47	2.52	1.72	1.44	1.72
19	5.32	1348	3.00	2.48	16.2	20	2.47	1.40	0.68	0.64
18	5.16	1358	3.00	2.43	16.8	40	2.45	1.42	0.69	0.65
39	5.97	1345	3.01	2.36	14.5	54	2.40	1.47	0.88	0.88
40	3.34	1358	3.00	2.66	35.8	58	2.54	1.31	0.43	0.42
57	7.07	1342	6.06	4.59	20.2	44	2.33	1.24	0.91	0.28
59	11.61	1350	11.17	8.6	22.6	26	2.43	1.23	0.8(est)	0.28
12	2.13	1600	1.05	0.87	14.4	50	2.46	1.51	0.95	1.52
42	3.30	1592	1.38	1.19	15.3	64	2.43	1.55	0.64	1.72
32	5.96	1601	2.98	2.50	15.1	16	2.41	1.51	0.69	1.52
29	5.92	1599	3.00	2.51	15.4	57	2.52	1.28	1.02	0.68
11	5.83	1591	3.01	2.74	23.9	44	2.46	1.21	0.71	0.48
31	7.93	1611	5.99	4.79	15.1	46	2.44	1.19	0.87	0.44
56	8.31	1590	6.07	4.71	14.8	51	2.50	1.19	0.88	0.44
58	3.64	1601	6.07	5.45	50.1	53	1.47	0.62	0.27	0.30
41	12.62	1601	11.16	8.48	15.9	28	2.45	1.12	0.68	0.30
60	12.75	1597	11.18	7.86	16.1	28	2.40	1.09	1.00	0.30

Table II

Recycle Gas Runs

Operating Data

Run No.	Dry Char Feed Rate Lbs./Hr.	Temp., °F	Press. Atm.	P_{H_2} Atm.	Gas Rate Total Recycle	SCF/Lb. Char Hydrogen Recycle	Res. Time Min.	% Sulfur Feed	% Sulfur Product	% H$_2$S in H$_2$ Obs. in Exit Gas	Equil.
27	3.69	1099	3.00	0.74	25.7	6.7	44	2.34	1.90	1.00	2.62
25	3.90	1330	1.34	0.72	14.6	8.5	52	2.50	1.76	1.07	1.73
38	1.60	1348	1.37	0.70	28.9	16.0	48	2.45	1.47	0.79	0.73
26	3.82	1345	3.01	1.38	26.3	12.8	52	2.37	1.48	1.07	0.76
33	1.70	1599	1.39	1.02	24.4	18.8	52	2.40	1.56	0.62	1.80
34	3.71	1594	3.01	2.12	23.7	17.3	57	2.44	1.28	1.03	0.60
36	5.49	1606	6.04	3.91	24.3	16.7	64	2.47	1.32	1.11	0.72

Table III

Yields and Material Balances -- Fresh Hydrogen Runs

Run No.	Balances Percent Closure				Yields -Wt.% Feed Char Dry			
	Total	Sulfur	Carbon	Hydrogen	Char	CH$_4$	CO	H$_2$O
54	-0.6	-15.3	-2.2	-7.4	87.9	2.8	2.4	2.3
55	+3.5	+3.3	-1.7	-4.8	-	3.5	3.3	2.3
51	+0.9	-7.6	-2.4	-7.4	87.2	4.0	2.8	1.8
14	-1.0	-1.1	-3.2	-2.0	87.0	2.7	3.7	4.1
16	-0.3	-6.6	-0.3	+0.5	87.4	3.8	3.1	2.8
23	-1.4	-8.6	-2.1	-6.5	88.5	3.7	3.2	2.3
19	-	-	-	-	-	4.6	2.3	3.3
18	-0.3	-1.6	-0.4	-0.9	85.5	6.0	2.6	3.3
39	-3.1	-2.8	-3.7	-4.6	-	6.4	2.4	3.4
40	-1.0	-8.2	-1.3	-3.5	84.1	7.0	4.7	2.9
57	+4.2	+3.0	+0.3	-6.5	-	8.5	4.7	3.4
59	-2.4	-8.5	-3.0	-	79.4	9.9	4.5	5.1
12	+2.6	+9.8	+4.9	-3.9	-	-	-	-
42	-2.7	-6.3	-0.4	-7.0	86.9	3.1	5.8	-
32	-1.3	-3.5	-2.9	+3.2	84.6	3.2	2.7	-
29	-3.3	+0.7	-2.6	-4.1	-	5.3	4.3	2.0
11	-3.0	+0.0	-1.3	-0.6	-	5.4	4.3	0.7
31	-0.5	-10.8	-0.7	-0.8	81.5	7.1	5.4	2.5
56	-3.7	-18.8	-8.4	-2.1	-	6.7	6.6	4.8
58	+7.3	-11.1	+2.8	-3.9	-	8.8	9.2	1.8
41	+2.1	-23.9	-1.7	-13.5	77.1	6.7	6.9	2.6
60	+30.8	+3.9	+31.5	-3.8	-	8.6	7.2	3.5

Table IV

Yields and Material Balances

Recycle Gas Runs

Run No.	Balances Percent Closure				Wt.% Yield		
	Total	Sulfur	Carbon	Hydrogen	Gas	Liquor	Solid
27	-3.6	-0.8	-2.6	+0.1	2.4	3.4	93.6
25	-6.9	-4.1	-4.6	-11.1	6.6	3.1	89.6
38	-0.4	+1.3	-0.7	+12.1	7.6	2.6	89.3
26	-6.3	+4.4	-6.0	+0.1	7.2	3.1	88.4
33	-13.1	-1.6	-11.8	-3.8	7.1	3.4	88.7
34	-6.2	+8.8	-5.9	-8.6	8.0	3.0	87.5
36	-0.9	+21.3	-1.7	-5.8	-	-	-

Table V

Gas Analysis and Yields

Recycle Gas Runs

Run No.	Gas Yield SCF/Lb.	Dry Gas Analysis						
		H_2S	CO_2	CO	CH_4	C_2H_6	N_2	H_2
27	0.64	0.26	11.7	16.6	43.7	0.4	1.9	25.5
25	2.51	0.63	3.6	11.6	24.3	0.3	2.2	57.4
38	2.72	0.44	3.5	14.7	24.7	-	2.9	53.8
26	2.42	0.52	2.3	14.8	32.0	0.3	2.7	47.4
33	3.94	0.48	1.4	10.1	10.6	-	2.4	75.0
34	4.06	0.75	0.8	12.9	11.8	-	2.5	71.3

Table VI

Solids Analysis - Recycle Gas Runs

Run No.	Ultimate Analysis (Dry Basis)					Sulfide Plus Pyritic Sulfur	Ash	Moisture
	H	C	N	O	S			
Average Feed Char	3.10	78.08	1.76	5.69	2.44	0.63	8.93	4.0
Product Chars								
27	2.72	82.07	1.77	1.93	1.90	0.18	9.61	
25	1.55	85.31	1.52	-0.17	1.75	0.23	10.04	
38	1.42	84.49	1.34	1.36	1.47	0.24	9.78	
26	1.90	85.11	1.45	-0.34	1.48	0.25	10.40	
33	1.06	85.57	1.21	-0.07	1.56	0.33	10.53	
34	0.78	85.15	1.15	0.98	1.28	0.38	10.66	

Figure 1

FLOW DIAGRAM OF CHAR DESULFURIZATION UNIT

FIGURE 2
EQUILIBRIUM DESULFURIZATION IN
CONTINUOUS FLUIDIZED SYSTEM

1 — 1100 °F - SINGLE STAGE

2 — 1350 °F - SINGLE STAGE

3 — 1600 °F - SINGLE STAGE

4 — 1600 °F - DEVOLATILIZATION
PLUS 1 STAGE

5 — 1600 °F - DEVOLATILIZATION PLUS 2
COUNTERCURRENT STAGES

6 — 1600 °F - IDEAL COUNTERCURRENT

FIGURE 3
VARIATION OF
DESULFURIZATION EFFICIENCY
WITH SEVERITY OF TREATMENT

FIGURE 4
APPROACH TO METHANE EQUILIBRIUM
IN FRESH HYDROGEN RUNS

FIGURE 5
APPROACH TO METHANE EQUILIBRIUM
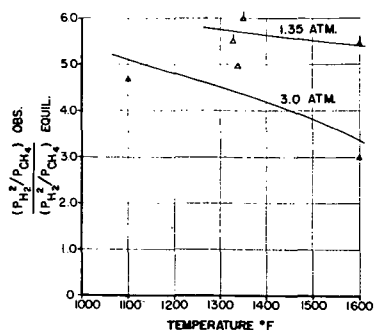IN RECYCLE GAS RUNS

FIGURE 6
FIRST ORDER
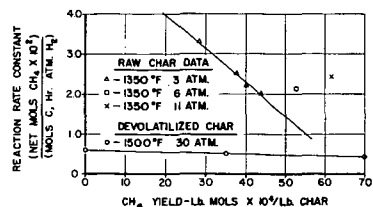REACTION RATE CONSTANTS FOR HYDROGENATION
OF FRESH AND DEVOLATILIZED CHARS

FIGURE 7
HYDROGEN YIELDS FROM CHAR
DEVOLATILIZATION